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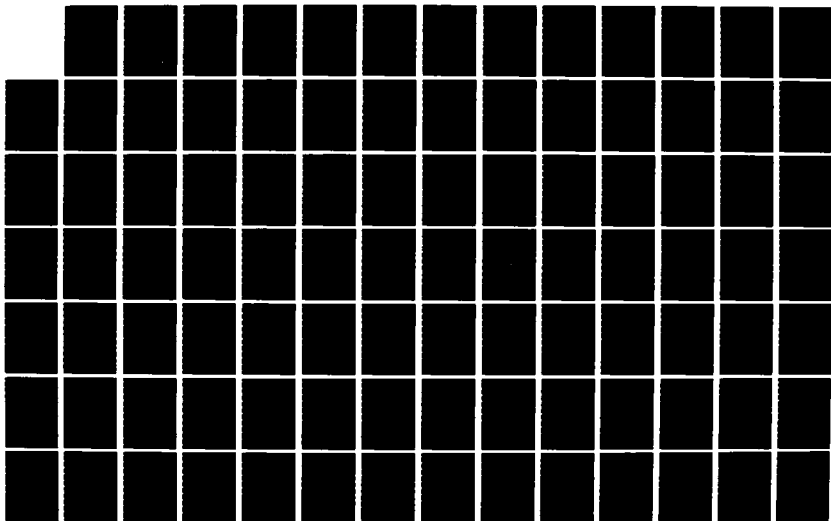
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COAL LIQUIDS: MANUFACTURE AND PROPERTIES—A REVIEW

INTERIM REPORT
AFLRL No. 162

By

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U.S. Army Fuels and Lubricants Research Laboratory
Southwest Research Institute
San Antonio, Texas

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20. ABSTRACT (Continue on reverse side if necessary and identify by block number) The review presented in the following pages represents an overview of the more prominent manufacturing and refining processes for generation of liquid products from coal. Some of the problems associated with the processes and products are discussed as well as methods for analysis and characterization. An annotated bibliography is included to provide the reader with immediately available literature references if more specific details are required.		

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FOREWORD

The review presented herein was conducted at the U.S. Army Fuels and Lubricants Research Laboratory (USAFRLRL), Southwest Research Institute, San Antonio, Texas under Contract No. DAAK70-82-C-0001 during the period October 1981 through September 1982. The work was funded, as part of a larger program, by the U.S. Army Mobility Equipment Research and Development Command (MERADCOM), currently the Belvoir Research and Development Center, Fort Belvoir, VA. Mr. F.W. Schaekel (STRBE-VF) served as contract monitor while Mr. M.E. LePera, Materials, Fuels and Lubricants Division (STRBE-VF) was the project technical monitor.

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INTRODUCTION

The conversion of coal to liquid products can be accomplished by two methods: direct liquefaction which converts coal directly to liquids, and indirect liquefaction by which the coal is gasified prior to formation of liquid products. Both methods are of German origin and were first developed into commercial-size operations between 1925 and 1935 by I. G. Farben and Ruhrchemie. The only method used today on a commercial scale is the indirect method developed by South African Coal, Oil and Gas (SASOL) in which the coal is initially gasified through a Lurgi refinement of the Fischer-Tropsch method.

Post World War II oil price increases and consequent fears of petroleum depletion in the near future awakened interest in coal conversion in the United States. However, as heated controversy took place over the comparative merits of the Fischer-Tropsch gasification method and direct liquefaction by hydrogenation, vast new petroleum reserves were discovered in the Middle East, thus bringing a halt to practical implementation of coal liquefaction technology.

For many years, coal has been a feedstock for chemicals used in both industry and medicine. As petroleum and natural gas sources become strained, the use of coal in the production of chemicals will expand. Methods by which coal can be converted to chemicals include those used in liquefaction processes: pyrolysis, hydrogenation, solvent extraction, and gasification followed by catalytic syntheses. Also catalytic oxidation with air is useful in the production of organic acids. Other methods are used for more selective chemical production such as hydrolysis, halogenation, and amination. In some cases, coal processing does not produce relatively pure chemicals ready for market, but rather produces solutions and mixtures with no clearly predominant product, so further refining is necessary.

Comparisons between prospective costs and revenue regarding different methods of coal liquefaction in the United States present a number of problems, and it is too early to tell if one process is better than another. All

processes discussed earlier are flexible, and no conclusions can be drawn until each has been tested on the demonstration scale. Some of the more prominent problems affecting the decision are those of cost estimations, severity of hydrogenation and cost of upgrading, operating costs, capital investment, competitiveness with petroleum prices, efficiency of conversion, quality and market value of products.

The purpose of this report is to describe each of the major coal liquefaction processes and stress the need to overcome inherent performance problems in the coal-derived materials, both independently and in relation to petroleum-derived fuels.

CHAR OIL ENERGY DEVELOPMENT (COED) PROCESS

History

FMC Corporation developed the COED process under the sponsorship of the Office of Coal Research and the U. S. Department of the Interior from 1962 through 1975. A process development unit was constructed in 1965 which could process 45 kg of coal per hour. A 33-ton/day pilot plant was put into operation and continued successfully until 1975, when federal funding was terminated by the Energy Research and Development Administration. FMC continued development privately thereafter. Current efforts are directed toward in-plant utilization of product gases.

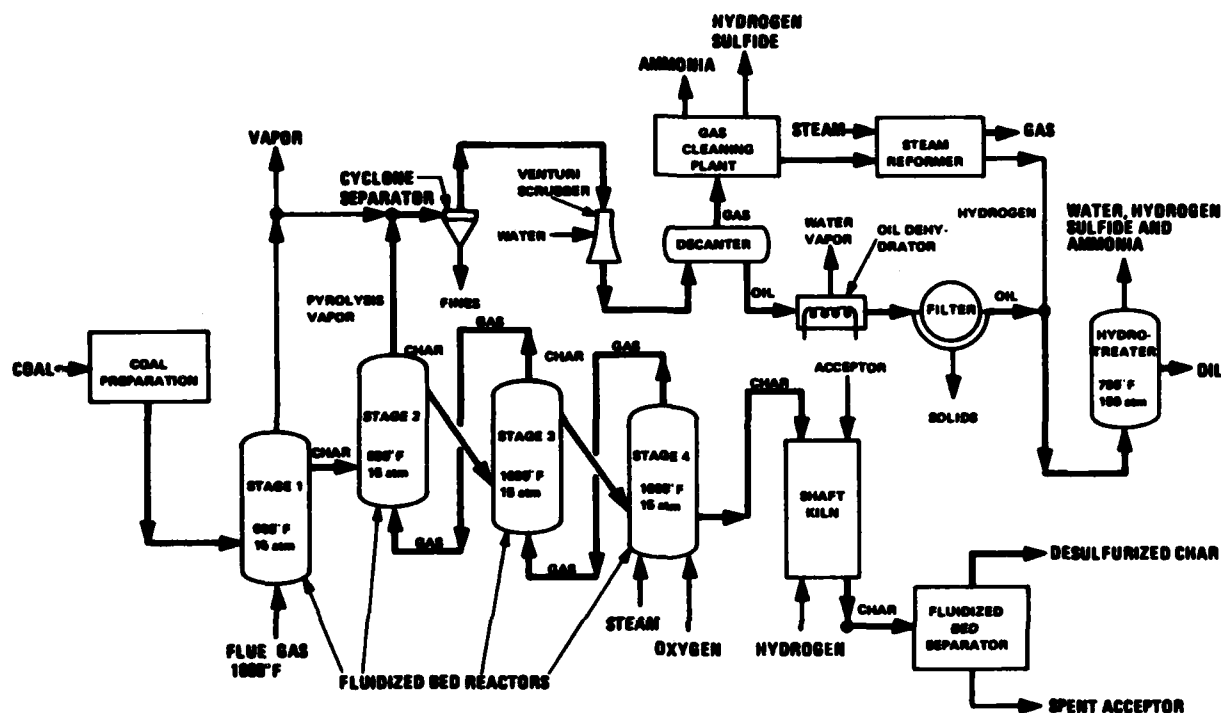
The greatest potential market for COED products is as a crude oil; however, several contingencies in the quality of the crude may have a bearing on consumer and legal acceptance. These aspects are outlined in the Problems section.(1, 2)*

The COED process is a pyrolysis rather than a hydrogenation process. The products are a low-sulfur crude oil and heat, which is used to generate electricity. Major steps in conversion are listed below.

1. Coal washing, pulverization, and drying
2. Pyrolysis
3. Separation of gases and oil vapors
4. Gasification of char and production of fuel gas
5. Recirculation of char and bed contents
6. Fuel gas recycle
7. Hydrocarbon gas production
8. Light oil hydrotreatment
9. Heavy oil filtration and hydrotreatment
10. Syncrude output

*Underscored numbers in parentheses refer to the list of references at the end of this report.

Heat is generated by the reaction of oxygen with a portion of char in the last pyrolysis stage. The number of stages in pyrolysis and the operating temperatures in each can be varied to accommodate coals with different physical properties. Figure 1 shows a schematic diagram of the plant process.



Reference: (43)

FIGURE 1. COED PROCESS SCHEMATIC

Properties

Table 1 summarizes properties for several COED crudes and cuts. Volatility of the Utah light fraction (blend B) and the Illinois coal fuel (blend F) correspond to that of a diesel fuel, while the distilled light fraction with heavy ends removed (blend G) corresponds to a gasoline. Upon separation, the aromatic content of COED crude tends toward the heavier fraction where it is least required. Sulfur and ash content are acceptable for all three cuts, but heat of combustion, viscosity, trace metal impurity, water and sediment, copper corrosion, and ignition characteristics do not meet speci-

fications. Thus, chemical composition of these fractions must be modified. Care must be taken when working with COED fuels because of low flash points and high polynuclear aromatic contents of the fractions. COED crudes and fractions are known to have an offensive odor as well as a high carcinogen content, thus adequate ventilation must be available when working with these materials.

Detroit Diesel Allison burned the Utah light, Utah heavy, and "Sea Coal" fluid (a topped crude product) obtained from the COED process in a T-63 combustor. Observations from testing were:

1. High NO_x emission probably due to fuel-bound nitrogen
2. High exhaust smoke due to high aromatic content
3. Flow difficulties due to high cloud point (70°-80°F)

Additional laboratory testing showed that:(3)

1. All samples had high trace metal content.
2. The light oil conformed to volatility limits for JP-4, but the heavier fraction had too many heavy components to serve as an acceptable marine diesel fuel.

Product Slate

Table 2 shows the projected COED product slate based on material balance for 10, 20, 30, and 40 thousand tons/day feed coal.(4)

Material and Energy Balance

Table 3 shows the material balance, and Table 4 shows the energy balance for the COED process.(4)

Operating Costs

Table 5 gives the cost of investment and operation of a full-scale COED processing unit based on 10, 20, 30, and 40 thousand tons/day feed coal input.(4)

**TABLE 1. COMPARISON OF COED FUELS WITH
SPECIFICATION PETROLEUM FUELS**

	A(1)	B	C	D	E	F	G	MIL-F-16884G DIESEL	MIL-F-24397 NAVY DISTILLATE	MIL-G-3056D MOTOR GASOLINE
Distillation (D 86), °F										
IBP	<300(2)	176	<300(2)	211	354	184	85			
10% off	<300	215	314	298	409	216	189		500 max	122-158
50% off	460	287	552	532	580	242	270		644 max	192-239
90% off	760	439	767	770	780	298	335	675 max	740 max	270-356
BP	~875	545	~849	---	---	340	382	725 max	---	---
Gravity, °API	18.4	41.9	22.5	22.2	18.4			---		
Saturates (FIA), vol%						88.5	68			
Olefins (FIA), vol%						2.0	10			
Aromatics (FIA), vol%						9.5	22			
Saturates (HPLC), vol%						84.7	77.6			
Olefins-Aromatics, (HPLC), vol%						22.4				
Induction Period, min				0.70	0.65	0.105	1440	0.30 max	---	
Total Acid Number, mg KOH/g						0.82	0.5			
Total Base Number, mg KOH/g							0.007			
Phenols, wt%							(0.002)			
Thiophenols, vol%							0.00001			
Free Sulfur, wt%						0.1	0			
Mono-aromatics (), vol%						10	11.4			
Di-aromatics (), wt%						0.3	0.3			
Tri-aromatics (), wt%						0.06	0.2			
Research Octane No.							74.3%			91.0
Motor Octane No.							66.2%			83.0
Cetane Number						24.3				
Elemental Analysis, wt%										
C							(82.48)			
H							(12.97)			
N	0.403	0.193	0.143				(0.12)			
S	0.02	0.01	0.05	0.28	0.16	0.112	(0.089)	1.00 max	0.30 max	0.10 max
O							(1.08)			
Cl							(3.36)			
F							(0.001)			
Total Aromatics, wt%	47	32	45							
Polynuclear Aromatics, vol%	12	3	14							
Aniline Point, °F	too dark	84.2	too dark	68	68					
Net Heat of Combustion (calc), Btu/lb.	17,782	18,356	18,020							
Ash, wt%	0.02	0.01	0.01	0.003	0.007	(0.001)	0.005		0.010	
Pour Point, °F	+55	<-65	+60	-10	+25		+20 max		+25 max	
Flash Point, °F	145	<+80	120	58	160		140 min		150 min	
Viscosity, cSt @										
80°F				6.43	12.20					
100°F	9.78	0.94	6.82	4.55	8.03		1.8-4.5		10.0 max	
160°F				2.25	3.36					
Trace Metals, ppm										
V	<5.3	<6.1	<5.4	0.1	0.1				0.5 max	
Na	3.42	0.92	6.13							
K	2.49	1.81	0.38							
Ca	12.7	<0.61	28.3							
Pb	<0.53	0.74	<0.54							
Mg	2.36	2.63	3.21							
Water and Sediment, wt%				0.50	0.10					
Copper Corrosion, 3 hr @ 210°F				1a	1a	4a	2b(1a)	1 max	2 max	1 max
Gross Heat of Combustion, (calc) Btu/lb.				18,980	18,800			19,500*	19,300*	

() = caustic washed

* = not specification requirements

(1) Blend Code: (Fuels derived by COED Process)

- A = U.S. Navy "sea coal" a mixture of Illinois and Utah coals
- B = Utah coal light fraction
- C = Utah coal heavy fraction
- D = raw fuel from FMC
- E = processed fuel from FMC

F = COED fuel from Illinois coal (as received)(AL-5294-F)

G = Light fraction of COED fuel with 26% heavy ends removed by distillation (AL-5236-F)

(2) BP distribution by gas chromatography

Reference: (3)

TABLE 2. COED PROCESS PRODUCT SLATE

<u>Output Material</u>	<u>Coal, Tons/Day (Moisture and Ash-Free)</u>			
	<u>10,000</u>	<u>20,000</u>	<u>30,000</u>	<u>40,000</u>
Oil, barrels/day	13,050	26,100	39,150	52,200
Power, MW	385	769	1,154	1,539
Sulfur, tons/day	392	784	1,176	1,568
Reference: (4)				

TABLE 3. COED MATERIAL BALANCE

<u>Input</u>		<u>Output</u>	
<u>Material</u>	<u>Wt Unit</u>	<u>Material</u>	<u>Wt Unit</u>
Coal	100.0	Products	
Water	470.0	Oil and electricity	12.3*
		Byproducts	
		Sulfur	2.4
		Ash and solid waste	33.2
		Cooling tower losses	439.0
		Gases to atmosphere	83.0
Total	570.0	Total	570.0

*This figure includes electrical power output at 556 kW-H/ton of coal input.

Reference: (4)

TABLE 4. COED ENERGY BALANCE

Input		Output	
Material	Higher Heating Value Units	Material	Higher Heating Value Units
Coal	100.0	Oil	27.5
		Electrical power	11.1
		Sulfur	1.1
		Cooling towers, exhaust	60.3
Total	100.0	Total	100.0
Reference: (4)			

TABLE 5. COED INVESTMENT AND ANNUAL OPERATING COSTS
(10⁶ dollars, 1st quarter 1978)

	Tons of Coal/Day			
	10,000	20,000	30,000	40,000
<u>Investment</u>				
Fixed capital	702	1,181	1,600	1,986
Total depreciable plant	830	1,396	1,892	2,349
Total capital required	876	1,473	1,997	2,478
<u>Annual Operating Costs</u>				
Materials and supplies	109.90	215.50	310.90	409.10
Labor	12.61	21.20	28.77	35.68
Property taxes & insurance	19.30	32.50	44.00	54.60
General & administrative overhead	2.16	4.04	5.84	7.60
Total Operating Costs	143.90	269.20	389.50	506.98

Reference: (4)

EXXON DONOR SOLVENT (EDS) PROCESS

History

The EDS process was the result of 10 years of coal liquefaction research by Exxon Corporation and was entirely self-funded. This work culminated with a one-ton/day pilot plant. Then, in 1976, Electric Power Research Institute and DOE, through cooperative agreement with Exxon, helped to construct a 250-ton/day pilot plant which was completed in 1980 with Exxon as the chief operator. Other participants now interested in this plant include Phillips Petroleum, ARCO, Ruhrkohle, and a Japanese company.

The Exxon Donor Solvent (EDS) Process (Figure 2) is similar to SRC-I and II (described later) except that the recycled oil for slurry is treated catalytically to restore its hydrogen donor character before being mixed with the coal.(2, 5) The following steps outline the liquefaction process:

1. Coal pulverization
2. Slurry preparation with hydrogenated recycle (donor) solvent
3. Hydrogenation
4. Distillation
5. Recycle (donor) solvent hydrogenation
6. Bottoms recycle

The heavy bottoms stream is fed to a flexicoking unit which, by low pressures and temperatures, produces high yields of liquids or gases from organic material in the feed coal.

The development of the market for light products is in Exxon's interest. The reason for this is that the proportion of light imported crude is diminishing. Thus, the proportion of fuel oil produced will eventually increase and require greater investment to produce neat gasoline diesel, jet fuel, and home heating oil. The result is that the cost of refining these products will rise and increase premiums over fuel oil. However, Exxon continues to research EDS fuel oil and possible markets for it.



Hydrogen requirements can be met from the process in three ways:

- GHLO2/A**

Properties

Middle and heavy distillate EDS products show greater differences from petroleum counterparts than do EDS naphthas. The oils are mostly aromatics (85 percent), some asphaltenes, and few saturates, whereas comparable petroleum fractions are better than 50 percent saturates. Volatility of the naphtha is almost identical to that of petroleum by ASTM D 86 distillation; however, its Reid vapor pressure is less than half that of petroleum. Other significant differences are higher existent gum, longer induction period, and lower octane number. Table 6 summarizes data for EDS Illinois No. 6 and Wyodak coals and compares their properties with those of petroleum unleaded gasoline.(2, 6)

Product Slate

Table 7 shows the product slate for EDS fractions.(5)

Typical boiling ranges and elemental compositions of heavy naphtha and fuel oil are given in Table 8.

The bottoms recycle feature of the EDS process provides liquefaction of some coal material that was not liquefied the first time through the system. This feature enables successful conversion of low-rank coals to liquids which would otherwise be difficult. Table 9 shows compositional data for Texas lignite.(7)

Operating Costs

An EDS pioneer plant on which the design of multiple commercial plants can be based is now in the planning stage. Table 10 gives the cost breakdown for such a facility.(5, 8)

TABLE 6. PROPERTIES OF NAPHTHAS FROM COAL

	Code Number, Description		
	G-P2 Petroleum Unleaded Gasoline	G-C2 EDS-Illinois No. 6	G-C3 EDS-Wyodak
Gravity, °API	58.2	39.5	36.2
Specific Gravity, 15.6/15.6°C	0.7459	0.8275	0.8438
Distillation, D 86, °C			
IBP	31	53	54
10% evaporated	53	94	97
20% evaporated	69	115	121
50% evaporated	107	148	160
90% evaporated	172	175	185
EP	199	187	198
Recovered, vol%	97.0	98	98.5
Residue, vol%	1.8	1	1.0
Loss, vol%	1.2	1	0.5
Reid Vapor Pressure, kPa (psi)	63.4(9.2)	22.7(3.3)	19.3(2.8)
Existent Gum, mg/100 ml,			
unwashed	1.1	145	258
washed	1.1	122	248
Oxidation Stability, minutes	1440	3180	3180
Copper Corrosion, rating	1A	4C	1A
Aromatics, vol%	29	30.8	32
Olefins, vol%	4	4.9	0
Carbon, wt%	85.81	85.25	85.14
Hydrogen, wt%	13.83	12.45	12.56
Nitrogen, wt%	0.05	0.223	0.207
Oxygen, wt%	0.18	1.92	1.84
Sulfur, wt%	0.003	0.61	0.21
Trace Metals, X-Ray	ND*	Fe	None
Motor Octane Number	83.6	73.9	77.6
Research Octane Number	91.0	78.7	83.8
Hydrogen-Carbon Ratio	1.92	1.74	1.76

*ND = Not determined.

Reference: (6)

TABLE 7. EDS SUGGESTED PRODUCT RANGE AND APPLICATION
(EXXON)

<u>Product</u>	<u>Percentage of Total Output</u>	<u>Typical Application</u>
C ₁ , C ₂ gas	20 max	Synthetic natural gas
C ₃ , C ₄ LPG	28 max	Convert to premium fuel
Naphtha	30-45	Gasoline blend stock
Middle distillate	22-30	Stationary turbine fuel
Heavy distillate	50 max	Low-sulfur fuel oil

Reference: (5)

TABLE 8. EDS PRODUCT ANALYSIS
70°F-C₇ REMOVED

	<u>Heavy Naphtha</u>		<u>200°C + Fuel Oil</u>	
	<u>Raw Liquid</u>	<u>Hydrotreated</u>	<u>Raw Liquid</u>	<u>Hydrotreated</u>
Distillation, (15/5), °C				
10 wt% off	106	92	247	239
50 wt% off	180	157	368	347
90 wt% off	199	182	433	412
Density, g/cm ³	0.87	0.8	1.08	1.01
Elemental Analysis, wt%				
C	85.60	86.80	89.40	90.80
H	10.90	12.90	7.70	8.60
O	2.82	0.23	1.83	0.32
N	0.21	0.06	0.66	0.24
S	0.47	0.005	0.41	0.04
Higher heating Value, MJ/kg	42.6	44.9	39.8	42.1

Reference: (8)

TABLE 9. COMPOSITION OF TEXAS LIGNITE

<u>Elemental Analysis, wt% DAF⁽¹⁾ Coal</u>	<u>Values</u>
Carbon	73.7
Hydrogen	5.5
Oxygen ⁽²⁾	17.8
Sulfur	1.7
Nitrogen	1.3
<u>Atomic Ratios, Organic Coal</u>	
H/C	0.86
O/C	0.20
<u>Gross Calorific Value, Btu/lb</u>	
MMF ⁽³⁾ Coal	9,500 ⁽⁴⁾
<u>Proximate Analysis, wt% Dry Coal</u>	
Volatile Matter	44.8
Fixed Carbon	41.9
Ash (SO ₂ -free)	13.3

(1) DAF = dry and ash-free
 (2) Oxygen determination by difference
 (3) MMF = moist, mineral-free
 (4) Shipment averages
 Reference: (7)

TABLE 10. OPERATING COSTS FOR EDS PIONEER PLANT

Capital Investment	\$2,800m
Stream days	310
Bbls of product/day	62,000
	\$/barrel
Operating cost	9.20
Coal	9.70
Capital charges	
(16% of investment)	<u>23.30</u>
	42.20
Reference: (5)	

H-COAL PROCESS

History

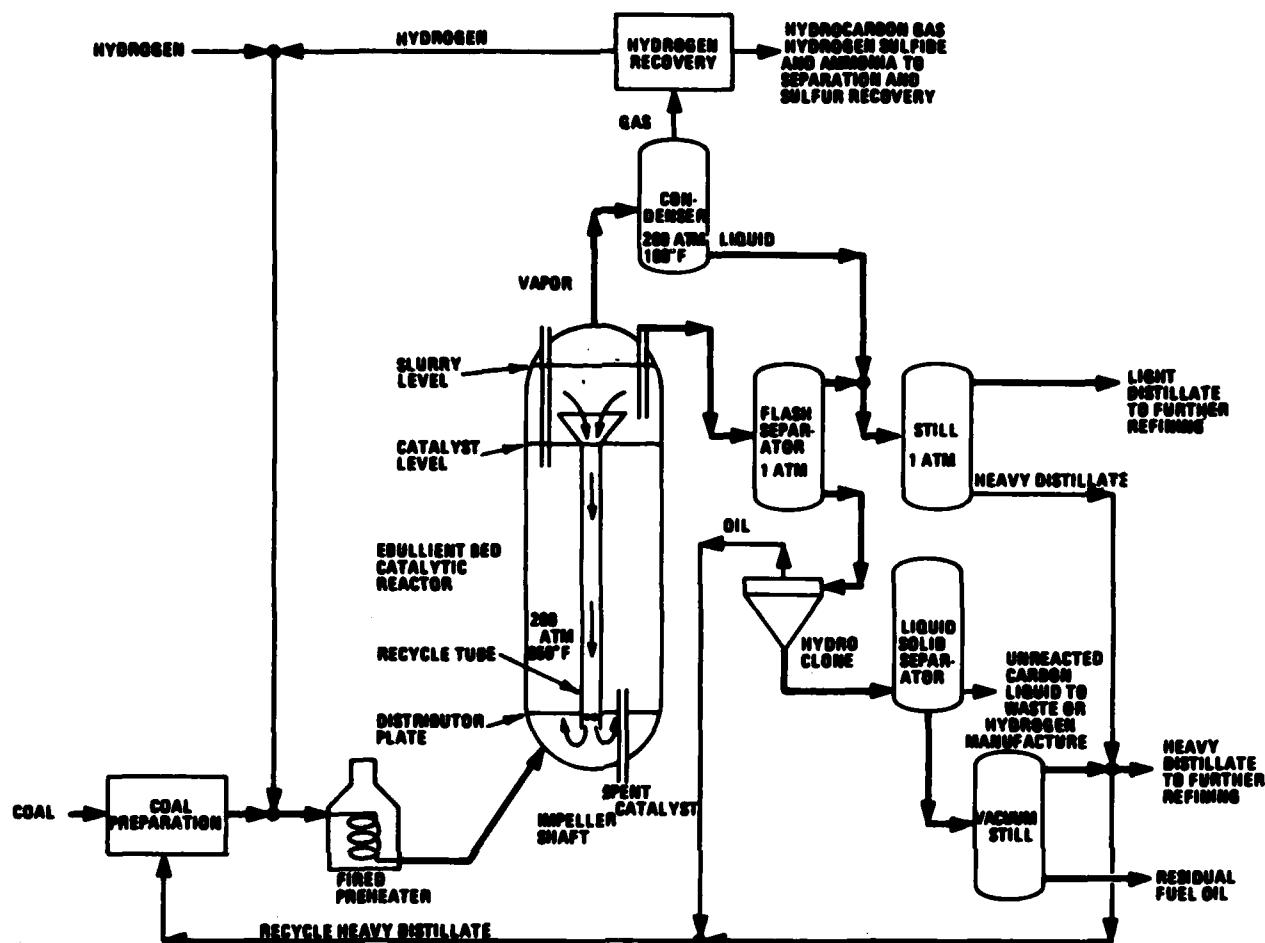
H-Coal was developed by Hydrocarbon Research, a subsidiary of Dynalectron. The process is an extension of HR's H-Oil process which desulfurizes heavy residues and converts them to lighter oils.

In 1962, DOE, Electric Power Research Institute, Ashland Synthetic Fuels, Conoco Coal Development, Mobil Oil, Standard Oil of Indiana, and the Commonwealth of Kentucky began with a bench-scale unit which evolved to a 3-ton/day unit. The same sponsors are now backing a 200- to 600-ton/day plant at Catlettsburg, Kentucky on which progress is under way. Ashland Oil, the operator, has estimated the feasibility of a 50,000-bbl/day plant on the basis of the pilot plant in Kentucky.

The H-coal process differs from other processes in that the coal-oil slurry comes in direct contact with a catalyst in the reactor. Five major steps are involved in the H-Coal process (shown in Figure 3):

1. Coal pulverization
2. Slurry preparation
3. Hydrogenation
4. Separation
5. Slurry oxidation/recycle and product collection

In the reactor, coal, recycle oil, and hydrogen react in the presence of a catalyst. Typical reaction temperatures and pressures can be varied to determine the net product yield and conform to product slate requirements. High severity yields an all-distillate material, while low severity yields a distillate and heavy fuel oil. Also, liquid flow from an ebullating pump causes the catalyst bed to expand and fluidize. Since the catalyst is in constant motion, a portion of it can be withdrawn and replaced at any time during reaction for the purpose of maintaining high catalyst activity.(10, 11)



Reference: (43)

FIGURE 3. H-COAL PROCESS

Product Slate

Table 11 shows a product slate for H-Coal process using Illinois No. 6 bituminous coal.

Breakdown of petrochemical feedstock production for a conceptual 50,000 bbl/day plant is given in Table 12.

Material Balance

Table 13 shows a material balance using an Illinois No. 6 bituminous coal.

Operating Costs

Cost of coal is similar for other liquefaction processes in the United States. Ashland Petroleum has an estimated investment at \$2.2 billion (March 1980 dollars). The cost breakdown is given in Table 14.

TABLE 11. H-COAL MATERIAL PRODUCT SLATE

Coal Feed	
Illinois No. 6	22,500 tons/day
Products	
C ₅ -360°F	16,000 bbl/day
Distillate Fuel Oil (360°-950°F)	24,600 bbl/day
Butane	3,800 bbl/day
Propane	<u>5,900 bbl/day</u>
TOTAL	50,300 bbl/day
Byproducts	
Sulfur	570 tons/day
Ammonia	160 tons/day
Synthetic Natural Gas	25 Mscf/day

Reference: (11)

TABLE 12. PETROCHEMICAL FEEDSTOCK PRODUCTION
(ASHLAND PETROLEUM COMPANY)

<u>Product</u>	<u>Quantity</u> (10^6 gal./yr)
Benzene	13.6
Toluene	24.5
Xylene	32.2
	(10^6 lb/yr)
Phenols and Cresols	170
Butane	222
Propane	205
Ethane	222

Reference: (10)

TABLE 13. H-COAL MATERIAL BALANCE

<u>Input</u>		<u>Output</u>	
<u>Material</u>	<u>Weight Units</u>	<u>Material</u>	<u>Weight Units</u>
Dry Coal	100.0	H ₂	(5.3)
		H ₂ O, CO, CO ₂	7.1
		H ₂ S, NH ₃	3.6
		C ₁ -C ₃	11.2
		C ₄ -400°F Naphtha	18.7
		400°-975°F	
		Fuel Oil	29.1
		975°F+, Bottoms	
		and Ash	<u>35.6</u>
Total	100.0	Total	100.0

Reference: (11)

TABLE 14. INVESTMENT COSTS FOR A 50,000 BBL/DAY
H-COAL PLANT (MARCH 1980 DOLLARS)

Capital Investment	\$2,200m
Stream days	310
Bbls of product/day	50,000
	\$ per barrel
Operating cost	16.80
Coal	11.20
Capital charges (16% on investment)	<u>22.70</u>
Total production cost	50.70

Reference: (5)

SOLVENT-REFINED COAL (SRC) PROCESS

SRC-I Process

History

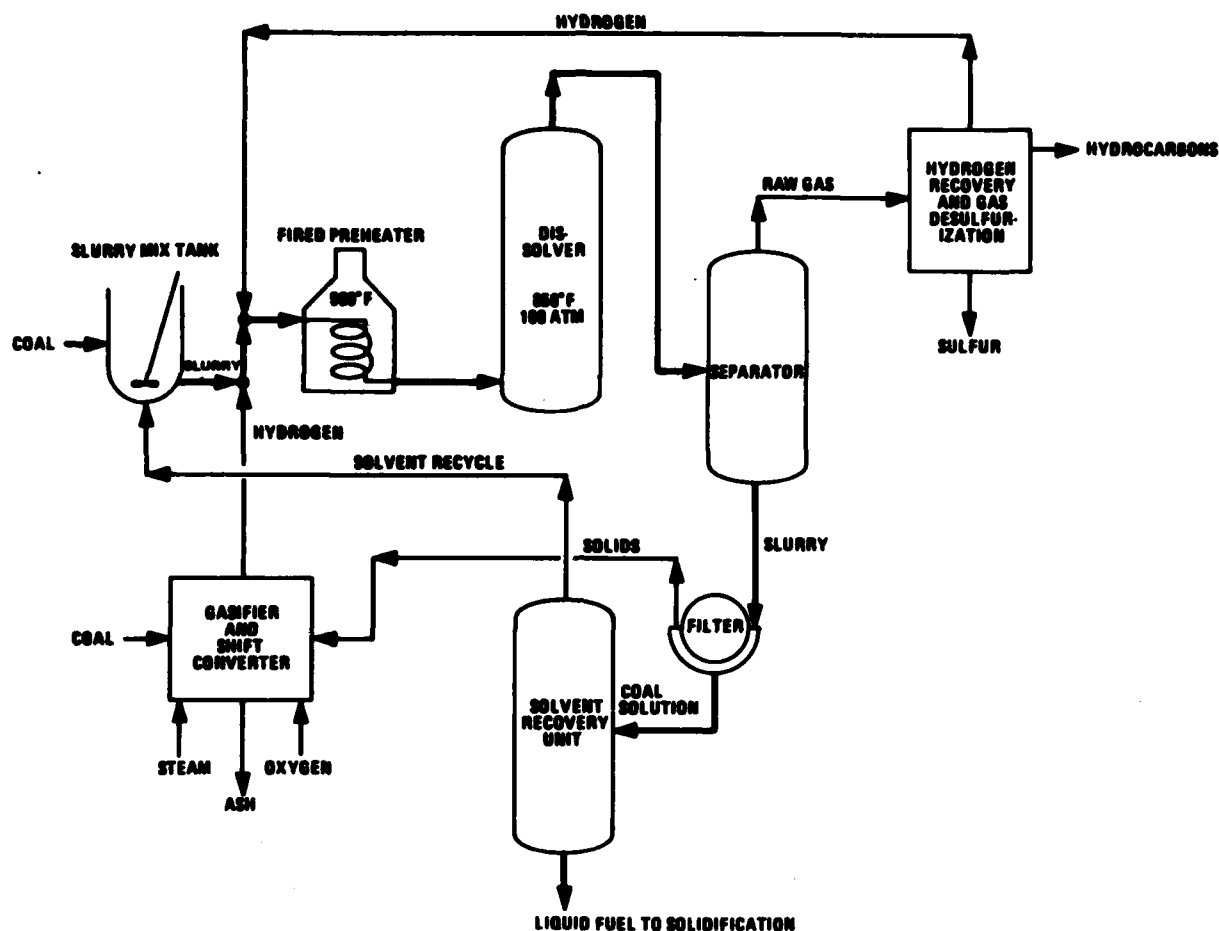
Initial development of this process in the United States was undertaken by the Spencer Chemical Company in the 1960's under a contract from the Office of Coal Research. Continued development was performed by the Pittsburgh and Midway Coal Mining Co. (a Gulf Oil subsidiary) which is currently building a 5400-ton/day demonstration plant, with joint funds from DOE, to be operational in 1984.

Pittsburgh and Midway Coal Mining Company began with a flow laboratory with a capacity of 50 lb/hr to experiment in the removal of sulfur and ash from coal. This led to the design of a 50-ton/day plant in 1969 by Stearns Rogers Corporation under Pittsburgh and Midway's direction. The U.S. Department of Energy helped funding, and the plant was completed in 1974. In 1977, the plant was modified for SRC-II.

Meanwhile, a small SRC-I plant (original process) was sponsored in 1972 by Edison Electric Institute, Southern Company, Alabama Power, Georgia Power, Gulf Power, and Washington Power. This program was coordinated with the SRC-II plant after 1978.

The SRC process has also been called the PAMCO process or P&M process. More recently, the full title of Pittsburgh and Midway Solvent-Refined Coal process has been applied.

The original Solvent-Refined Coal Process (SRC-I) (Figure 4) was developed to convert high-sulfur coal to a commercially acceptable solid fuel. About 70 percent of the total output product is a solid "coal," and about 9 percent is a distillate liquid. A two-stage liquefaction process for SRC-I was also conceived which would increase the liquid yield. The major steps in the SRC-I process are as follows:



Reference: (43)

FIGURE 4. SRC-I PROCESS

1. Coal pulverization
2. Slurry preparation
3. Coal dissolution
4. Hydrogenation
5. Distillation
6. Solid removal

Three streams are produced in the fifth step: gas, hydrocarbon coordinates (liquid), and the product slurry from which the solid refined coal is obtained (Step 6).

Product Slate

Table 15 shows the products and byproducts for SRC-I.(5)

Material and Energy Balance

Table 16 shows the material and energy balance for a SRC-I process.(5)

Two-Stage Liquefaction for SRC-I

A tight market for SRC solid spurred an investigation of how to upgrade the products to create more liquid fuels. The study concluded that a two-stage liquefaction process followed by catalytic hydrocracking of the SRC product could be more economical than the SRC-II process. Marked reduction in sulfur and nitrogen in the SRC solid and a greater production of fuel oil and naphtha were forecast. Different investments and their effect on costs are given in Table 17.

TABLE 15. PRODUCTION PER STREAM DAY (excluding sulfur)

	<u>Demonstration</u>		<u>Two-Stage</u>		<u>Two-Stage</u> <u>Minus Convention</u>
	<u>Plant Design</u>		<u>Liquefaction</u>		
	<u>Tons</u>	<u>% of Total</u>	<u>Tons</u>	<u>% of Total</u>	
Fuel gas	Nil	-	670	3.4	+ 670
Naphtha	1,370	6.7	2,430	12.3	+1060
Fuel Oil	3,210	15.7	8,610	43.7	+5400
SRC Solid	15,820	77.6	7,990	40.6	-7830
TOTAL	20,400	100.0	19,700	100.0	- 700
Reference:	(5)				

TABLE 16. MATERIAL AND ENERGY BALANCES FOR INPUT/OUTPUT OF SRC-I
FOR A CONCEPTUAL COMMERCIAL PLANT

	<u>Input/Output</u>	<u>% of Total</u>
Feedstock, tons/stream day		
Coal feed to dissolver	30,000	
Coal feed to gasifier	2,825	
Total Coal by weight	<u>32,825</u>	
Total Coal, million Btu/hr	35,015	
Products and Byproducts,		
Tons/stream day		
Solid SRC (850°F+)	15,712	73.7
Sulfur	957	4.5
Naphtha (C ₅ -30°F)	1,461	6.8
Distillate ⁵ (350°-850°F)	<u>3,210</u>	<u>15.0</u>
Total Products and Byproducts	21,230	100.0
Products and Byproducts, Million Btu/hr		
Solid SRC	20,680	74.7
Sulfur		
Naphtha	2,331	8.4
Distillate	<u>4,662</u>	<u>16.8</u>
Total Products and Byproducts	27,682	100.0
Thermal Efficiency = 73.4		

Reference: (5)

TABLE 17. COST COMPARISON OF MODIFICATIONS

Capital Investment	\$1,870m
Stream Days	330
Bbls of product/day	110,000
Operating cost	\$/barrel
Coal	5.40
Capital charges	10.50
(16% of investment)	9.10
Total production cost	25.00

Reference: (5)

SRC-II Process

History

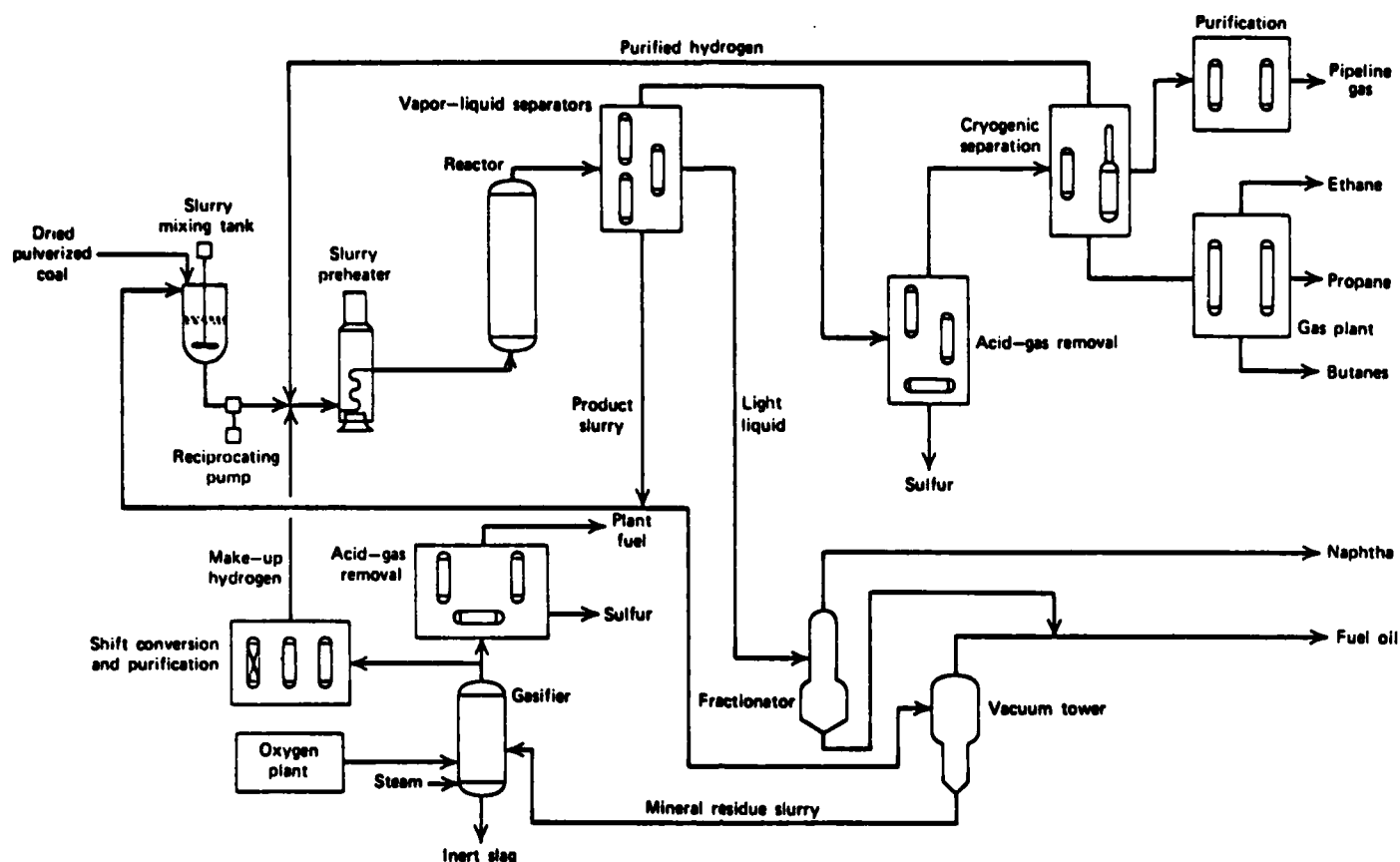
In 1979, Gulf Oil Company analyzed a market which SRC-II may be able to penetrate because of technical benefits and reduced capital and operating costs compared to other fuels. The competition is mainly with high-sulfur coal, and wherever small boilers operate at a low-load factor such as in manufacturing plants. The major market is for fuel oil, the largest SRC-II product. The analysis did not mention any side benefits for the consumer; thus, it is assumed that market penetration will depend upon trends in oil prices.(5)

Gulf Oil holds some patents for the SRC-II process for a planned demonstration plant, although those acquired as a result of DOE-funded operations can be made available by the government to qualified United States applicants. Funding of the demonstration plant is to be 50 percent by DOE and 25 percent each by the German and Japanese government.(13)

The SRC-II process (Figure 5) differs from SRC-I in that the coal-solvent slurry is recycled as the feed stream rather than using fresh distillate material. Also, a greater proportion of coal is converted to liquid and gas which can be distilled, thereby eliminating the problem of separating solids from the remaining product. Five major steps are involved in the conversion of raw coal to finished products by the SRC-II process:

1. Coal pulverization
2. Slurry preparation
3. Hydrogenation
4. Distillation
5. Slurry recycle and product collection

After reaction (Step 3), the process gas is cooled while H_2S and CO_2 are removed. Further separation of H_2 from gaseous hydrocarbons allows fumes to be recycled to become product materials.



Reference: (13)

FIGURE 5. SRC-II PROCESS

Properties

When compared with military specifications, the hydrotreated SRC-II naphtha falls between that of Type I, Grade B and Type II petroleum aromatic naphtha as shown by specific gravity and distillation data.

SRC-II fuel oil has a lower viscosity and contains less sediment and sulfur than either petroleum No. 4 or No. 6 burner fuels. However, it has a slightly higher nitrogen content and a higher density, revealing that it has a lower heat of combustion per unit weight than do the petroleum fuels.

The SRC-II middle distillate compares well with specifications for petroleum No. 2 burner fuel, the only differences being slightly higher specific gravity and viscosity.

Tables 18, 19, and 20 show data for SRC-II naphtha fuel oil and distillates, and specification for corresponding petroleum cuts.

Table 21 gives the approximate projected composition of saleable pipeline gas from the SRC-II process.

Product Slate

Table 22 shows the projected SRC-II product slate based on material balance for 10, 20, 30, and 40 thousand tons/day feed coal.

Material and Energy Balance

Table 23 shows the input/output material balance for the SRC-II process. The figures are based on coal and water input of 100 units and 123 units by weight, respectively.

Table 24 shows the energy balance visualized for a future commercial plant.

Operating Costs

Investment costs for start-up and operating costs for a commercial SRC-II plant are given in Table 25.

TABLE 18. PROPERTIES OF SRC-II FUEL OIL COMPARED TO PETROLEUM SPECIFICATIONS

	SRC II FUEL OIL*	PETROLEUM NO. 4*		PETROLEUM NO. 6*		SPEC. FOR NO. 4 LIGHT		SPEC. FOR NO. 4		SPEC. FOR NO. 6	
Gravity, °API	8.3	23	12			30 max					
Viscosity, @ 100°F, cSt	4	14-20	900			2.0-5.8		5.8-26.4			
Flash Point, °F	>150	>150	>150			100		130		140	
Pour Point, °F	-23	<-20	<460			20		20		15	
Sediment, wt%	<0.03	0.05	0.25								
Ash, wt%	0.015	0.01	0.05			0.05		0.10		1.00	
Nitrogen, wt%	0.9	0.2	0.4			---		---		---	
Sulfur, wt%	0.25	1.00	0.3-2.2			---		---		---	
Net Heating Value, Btu/gal	148,000	135,000	141,000			report		report		report	
Carbon Residue, wt%	<0.3	5.5	6-16			---		---		---	

Reference: (13)

TABLE 19. PROPERTIES OF SRC-II DISTILLATES

	MIDDLE DISTILLATE		BURNER FUEL NO. 2		HEAVY DISTILLATE	
Specific Gravity, 60°/60°F	0.98		0.88		1.08	
Viscosity @ 100°F, SUS (@, 200°F)	38		32.6-37.9		(40)	
Pour Point, °F	-50		20 max		+50	
Flash Point °F	170		100 min		300	
Nitrogen, wt%	0.8		---		1.1	
Sulfur, wt%	0.2-0.25		1.0 max		0.3-0.4	
High Heating Value, Btu/lb	17,400				app. 17,000	
Distillation, °F						
1BF	370		---		580	
10% off	390		---		610	
50% off	470		---		690	
90% off	570		540-640		800	
EP	600		---		900	

Reference: (13)

TABLE 20. PROPERTIES OF SRC-II NAPHTHA COMPARED TO PETROLEUM SPECIFICATIONS

	SRC-II (1)	SPEC. FOR SOLVENT NAPHTHA (2)		SPEC. FOR ALIPHATIC NAPHTHA (3)	SPEC. FOR AROMATIC NAPHTHA (4)		
		TYPES A and B	TYPE C		GRADE A	GRADE B	TYPE II
SPECIFIC GRAVITY @ 20°/20°C @ 15.6/15.6°C	0.780	0.840-0.885	0.915-0.945	0.708-0.768	0.810-0.871	0.770-0.847	0.825-0.875
DISTILLATION, °C vol% off							
1BP	68			85 max	88 min	88 min	129 min
10	90		160 max	102 max			171 min
30	132						
50	149			107 max	100-116	100-166	143-168
70	160		187 min				182-200
90	174	160 max		121 max			
EP	193	180 min		143 max	140 max	140 max	218 max
NITROGEN, ppm	4500 (5)						
Unhydrotreated							
Hydrotreated	0.2						
SULFUR, ppm	1900						
Unhydrotreated							
Hydrotreated	0.5						
OXYGEN, ppm	3.5						
Unhydrotreated							
Hydrotreated							
AROMATICS, vol%	34						
Unhydrotreated	14						
Hydrotreated							
NAPHTHENES	45						
Unhydrotreated	62						
Hydrotreated							
Paraffins	21						
Unhydrotreated	24						
Hydrotreated							

- (1) Reference: (4)
 (2) MIL-N-15178B, 5 January 1968.
 (3) TT-N-95b, July 10, 1967.
 (4) TT-N-97F, May 28, 1962.
 (5) Reference: (13)

TABLE 21. PROJECTED SRC-II SNG COMPOSITION

<u>Component (dry basis)</u>	<u>Vol%</u>
Methane	84.27
Ethane	8.61
Propane	1.69
Nitrogen	3.34
Hydrogen	1.54
Carbon monoxide	0.09
Carbon dioxide	<u>0.46</u>
TOTAL	100.00

Reference: (4)

TABLE 22. SRC-II PROCESS PRODUCT SLATE

<u>Output Material</u>	<u>Feed Coal, Tons/day,</u> <u>(Moisture & Ash-free Basis)</u>			
	<u>10,000</u>	<u>20,000</u>	<u>30,000</u>	<u>40,000</u>
Synthetic natural gas, SCF/day	54.36	108.7	163.1	217.4
Propane, bbl/day	1,980	3,958	5,932	7,916
Butane, bbl/day	1,337	2,677	4,810	5,351
Naphtha, bbl/day	3,075	6,151	9,225	12,305
Fuel Oil, bbl/day	18,477	36,957	55,429	73,915
Sulfur, tons/day	404.5	809	1,212	1,616
Ammonia, tons/day	29.5	58.8	88.5	118.1

Reference: (4)

TABLE 23. SRC-II MATERIAL BALANCE

Input		Output	
<u>Material</u>	<u>Weight Units</u>	<u>Material</u>	<u>Weight Units</u>
Coal	100.0	Synthetic natural gas	8.4
Water	<u>123.0</u>	Propane	1.1
		Butane	0.9
		Naphtha	2.7
		Oil	24.1
		Sulfur	2.7
		Ammonia	0.2
		Slag	8.9
		Gases to Atmosphere	49.0
		Water Losses	<u>125.0</u>
TOTAL	223.0	TOTAL	223.0
Reference: (4)			

TABLE 24. SRC-II ENERGY BALANCE

<u>Form of Energy</u>	<u>Pounds/Hr</u>	<u>Million Btu/Hr (HHV)</u>
<u>Input</u>		
Coal	2,791,000	35,775
Electricity	10,430 kW	99
Total Input		<u>35,874</u>
<u>Output</u>		
Main Products		
Fuel Oil	811,820	13,806
Pipeline Gas	86,580	2,002
Naphtha	217,400	3,974
Light Hydrocarbons	240,410	5,291
Butanes	25,163	526
Total Main Products		<u>25,599</u>
Byproducts		
Sulfur	98,417	391
Ammonia	15,292	150
Tar Acids	3,100	65
Total Byproducts		<u>606</u>
Total Output		26,205
Thermal Efficiency		
Main Products Only		71.4%
Main Products Plus Byproducts		73.0%
Reference: (14)		

TABLE 25. COSTS FOR A CONCEPTUAL COMMERCIAL SRC-II PLANT

Capital Investment	\$1,983m
Stream Days	330
Bbls of product/day	85,000
Operating cost	\$/barrel
Coal	3.90
Capital charges	
(16% of investment)	13.20
Total production cost	11.40
Reference (5)	

FISCHER-TROPSCH PROCESS

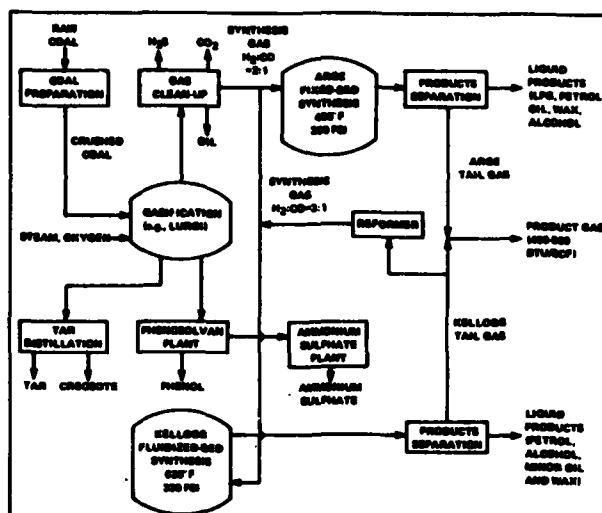
History

The United States Bureau of Mines studied Fischer-Tropsch synthesis in pilot plants in Pennsylvania and Missouri during the late 1940s and early 1950s. As interest in the process is now increasing, the Department of Energy has begun a new R&D program, and construction of a new unit to investigate process modifications has begun.

Since Fischer-Tropsch products are very similar to petroleum products, consumer demand is essentially the same. After design problems are overcome and United States coals accommodated, the only difference will be cost effectiveness for the Fischer-Tropsch process versus the same for petroleum processing.

The Fischer-Tropsch Process, unlike other coal liquefaction methods, is an indirect process in which the coal is first gasified and the gases are reacted to form liquid products (Figure 6). Although Fischer-Tropsch is the most highly-developed liquefaction process due to the efforts of South Africa and the South African Coal, Oil and Gas Corp. (SASOL) starting in the 1950's, its technology in the United States is not economically justifiable thus far for the following reasons:

1. The gasification unit used at SASOL is based on Lurgi (gasification) technology and cannot process the agglomerating classes of coal abundant in the United States;
2. The gasoline product has a low octane rating and would require upgrading;
3. The paraffin waxes produced have commercial value in South Africa but not in the United States;
4. The hydrocarbon gas produced cannot be used directly by United States consumers. It cannot be considered a synthetic natural gas due to its low heating value.



Reference: (43)

FIGURE 6. FISCHER-TROPSCH PROCESS

This section outlines a United States version of a Fischer-Tropsch plant designed with major modifications as a result of a DOE economic analysis.(4, 15)

Fischer-Tropsch liquefaction is complex when compared to other processes. The major steps in conversion are as follows:

1. Coal pulverization
2. Delivery to ash removal and gasifier
3. Steam and oxygen addition
4. Gas production
5. Gas to heat exchanger and shift converter
6. Separation of tar, sulfur, carbon dioxide, and acid gas
7. Liquefaction
8. Heavy-light hydrocarbon separation
9. Tail gas reforming

10. Light oil, gas, and hydrogen separation
11. Hydrogen recycle
12. Light hydrocarbon isomerization and cracking

Liquefaction occurs in circulating catalytic fluidized bed reactors, and product composition is dependent on operating conditions. The Fischer-Tropsch reaction is quite exothermic, and the resultant heat is used to raise steam for other units.

Properties

Since Fischer-Tropsch is well-developed and has been used successfully by SASOL, intensive research has been done on existent properties of liquid products for application of a conceptual commercial plant in the United States. Table 26 lists prospective chemical composition of Fischer-Tropsch liquefied petroleum gas.

Table 27 summarizes properties of naphtha, diesel, and fuel oil from Fischer-Tropsch liquefaction.

Table 28 summarizes the composition of synthetic natural gas.

Products

Table 29 lists product and byproduct yield of a conceptual Fischer-Tropsch facility.

Table 30 shows the conceptual saleable product yield based on a 10.95 million tons of coal per year input.

Material and Energy Balance

The Fischer-Tropsch primary material balance for the process is given in Table 31.

The primary energy balance based on percent energy of coal input by higher heating value and on energy of product output at 100-percent capacity of operating facility is given in Tables 32 and 33. In addition, 2.97×10^6 kW-hr/day of electrical power are produced.

TABLE 26. FISCHER-TROPSCH LPG COMPOSITION

<u>Component</u>	<u>Wt%</u>	<u>Vol%</u>
Butylene	16.2	15.8
Butane	82.1	82.6
Propane-Propylene	0.1	0.1
Pentanes	1.7	1.6
Sulfur	Nil	----
Nitrogen	Nil	----

Reference: (4)

TABLE 27. PROPERTIES OF SOME FISCHER-TROPSCH PRODUCTS COMPARED TO PETROLEUM SPECIFICATIONS

	BUTANE	LIGHT NAPHTHA	HEAVY NAPHTHA	DIESEL FUEL	FUEL OIL	BUTANE SPECIFICATION ASTM D 1835	DIESEL FUEL SPECIFICATION ASTM D 975 (#1-D)	FUEL OIL SPECIFICATION ASTM D 396 (#1)	GAS TURBINE FUEL OIL SPEC. ASTM D 2880 (3-6T)
Vapor Pressure @ 100°F, psig	37					70 max			
Volatile Residue Evaporated Temperature 95%, °F	31					36 max 2.0 max			
Pentane and Heavier, %	1.6								
Residual Matter	0.05								
Residue on Evaporation 100 ml	0					0.05 max 15 max			
Sulfur, grains/ft ²									
Sulfur, wt%				NIL	NIL		0.5 max	0.5 max	---
Higher Heating Value, Btu/lb.		20,815	20,430						---
Gravity, °API		85.5	71.3	57	19,855		---	35 min	---
Cetane Number		plus 30	plus 30	60+	(Dark)		40 min	---	---
Saybolt Color		0.8	1.7						
Viscosity @ -30°F, cSt				1.4	115		1.4-2.5	1.4-2.2 150-300	45 min
Viscosity @ 100°F, cSt									
Viscosity @ 100°F, SUS									
Distillation (ASTM D 86), °F									
1BP		96	186	301	572		---	420 max	---
10Z		115	208	327	665				
30Z		128	226	351	715				
50Z		137	236	372	759				
70Z		146	253	421	811				
90Z		159	266	539	873				
EP		185	300	637	885		550 max	550 max	---
Freezing Point, °F		-136	-67						---
Pour Point, °F				10	150				---
Aniline Point, °F		140	160	175					
Oxygen, wt%		0.2	0.5	0.5	0.5				
Ash, wt%				0.01					
Nitrogen, ppm				NIL			0.01 max	---	0.03 max
Neutralization No., mg KOH/g				0.1 25					
Sapon Point, mm									
Copper Corrosion @ 150°F	1	1	1	1		1 max	3 max	3 max	
@ 210°F	1	1	1	1		1 max	3 max	3 max	
Water and Sediment, %						Trace	Trace	Trace	1.0 max
Flash Point, °F				100	300	100 min	200 min	200 min	130 min

Reference: (4)

TABLE 28. FISCHER-TROPSCH SNG COMPOSITION

<u>Component</u>	<u>Vol%</u>
Hydrogen	1.1
Carbon monoxide	0.1
Carbon dioxide	1.4
Nitrogen	6.8
Methane	83.7
Ethane	1.5
Propane	1.8
Butane	<u>3.6</u>
TOTAL	100.0

Reference: (4)

TABLE 29. FISCHER-TROPSCH PRODUCT YIELD

<u>Products</u>	<u>Tons/Stream Day</u>	<u>%</u>
Unleaded gasoline	6,010	58
Diesel fuel	1,055	10
Ethylene	865	8
Alcohols	400	4
<u>Byproducts</u>		
Tars	840	8
Ammonia	195	2
Sulfur	<u>1,015</u>	<u>10</u>
TOTAL	10,380	100

Reference: (15)

TABLE 30. FISCHER-TROPSCH PRODUCT SLATE

<u>Product</u>	<u>10⁶ ton/year</u>	<u>Wt%</u>
Motor fuels	2.15	64
Ethylene	0.32	10
Tar products	0.28	8
Ammonia	0.07	2
Sulfur	0.37	11
Chemicals	0.15	5
TOTAL	3.34	100
Reference: (15)		

TABLE 31. FISCHER-TROPSCH MATERIAL BALANCE

<u>Input</u>		<u>Output</u>	
<u>Material</u>	<u>Weight Units</u>	<u>Material</u>	<u>Weight Units</u>
Coal	100.00	Synthetic natural	
Water	180.00	gas	16.48
		Butane	0.85
		Naphtha	5.95
		Oxygenates	1.14
		Diesel	5.26
		Fuel oil	1.79
		Sulfur	2.54
		Solid waste, ash	9.00
		Vent gases	106.00
		Cooling Tower Losses	130.99
TOTAL	280.00	TOTAL	280.00
Reference: (4)			

TABLE 32. FISCHER-TROPSCH ENERGY BALANCE

Input		Output	
Material	Energy Units	Material	Energy Units
Coal	100.00	Synthetic natural gas	35.56
		Butane	1.91
		Naphthas	13.02
		Oxygenates	1.52
		Diesel oil	11.32
		Fuel oil	3.76
		Sulfur	1.07
		Electrical power	1.52
		Cooling towers, etc.	30.32
TOTAL	100.00	TOTAL	100.00

Thermal Efficiency = 69.0%

Reference: (5)

TABLE 33. ENERGY BALANCE OF A FISCHER-TROPSCH FACILITY
OPERATING AT 100 PERCENT CAPACITY
(Based on coal input of 10.95×10^6 tons/year and 12,550 Btu/lb HHU)

Product	Output	Heating Value	% of Coal Feed
Liquid Products			
Regular gasoline	781×10^6 gal./yr	125,000 Btu/gal	35.5
Diesel fuel	130×10^6 gal./yr	120,000 Btu/gal	5.7
Ethylene	317,000 tons/yr	4.0×10^7 Btu/ton	4.6
Alcohols	145,000 tons/yr	2.5×10^7 Btu/ton	1.3
Byproducts			
Tar products	307,300 tons/yr	4.0×10^7 Btu/ton	4.5
Ammonia	71,700 tons/yr	2.4×10^7 Btu/ton	0.6
Sulfur	370,000 tons/yr	8.0×10^6 Btu/ton	1.1
Electricity	$1,083 \times 10^6$ hr/yr	3,413 Btu/W-hr	1.3
TOTAL			54.6

Reference: (15)

SYNTHOIL PROCESS

History

The Synthoil process (Figure 7), developed by the U.S. Bureau of Mines at the Pittsburgh Energy Research Center, is a direct hydrogenation process in which the oil portion of the coal/oil slurry is a vehicle for moving coal. It is similar in this respect to the H-COAL and COED processes, but differs from the EDS + SRC (solvent extraction) process in which the recycle liquid is prehydrogenated and acts as a hydrogen source for the coal.(4)

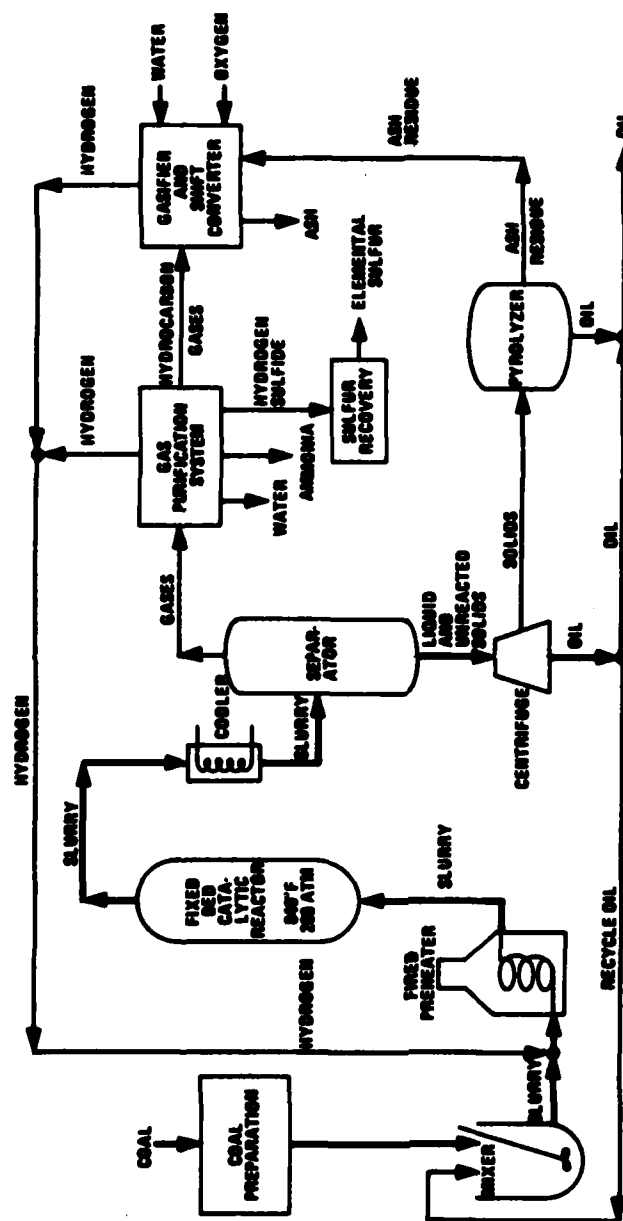
In 1974, the BuMines Evaluation Group published an economic study of a projected Synthoil facility designed to produce 108,000 bbl/day of crude from western Kentucky coal. At that time, a 23 kg/hour experimental system was in operation and a 9- to 14-ton/day pilot plant had been designed which was constructed in 1975.

The Department of Energy stopped active involvement in the Synthoil process in 1978 due to an unresolved process problem.(2)

The following basic steps outline the Synthoil process:

1. Coal pulverization
2. Slurry preparation
3. Hydrogenation
4. Phase separation and distillation
5. Liquid fuel oil production
6. NH_3 , H_2S , H_2O removal from gaseous stream
7. Hydrogen recycle
8. Pyrolysis of solids

After pyrolysis, resulting residue and hydrocarbon gases are used to make recycle hydrogen. Some coal can be added directly to the gasifier to increase H_2 yield. Ash is disposed of as mine fill, and NH_3 and H_2S (after conversion to elemental sulfur) become byproducts.(9)



Reference: (43)

FIGURE 7. SYNTHOIL PROCESS

Properties

Properties of finished products vary according to feed coal characteristics, although to some extent they are similar. Table 34 gives some physical properties and composition for three cuts of West Virginia coal liquid product.

Product Slate

Table 35 shows a projected product slate for plant sizes processing 10, 20, 30, and 40 thousand tons/day of feed coal.

Tables 36 and 37 show the Synthoil product breakdown in terms of volatility distribution of the overall product and hydrocarbon-type composition of a middle and heavy distillate from West Virginia coal.

Material and Energy Balance

Table 38 shows the material balance for the Synthoil process.

Tables 39 and 40 show thermal balance for the Synthoil process at two different times by two different sources.

Operating Costs

Tables 41 and 42 show required investment and projected annual operating costs for a Synthoil process plant utilizing 10, 20, 30, and 40 thousand tons of feed coal per day.

TABLE 34. PHYSICAL PROPERTIES OF SYNTHOIL LIQUIDS

	Distillate Type			MIL-F-24397 Heavy Distillate
	Light 207°C	Middle 207°-363°C	Heavy 363°-531°C	
Specific Gravity 60°/60°F	0.936	0.990	1.109	
Gravity, °API	19.7	11.4	---	27 min
Pour Point, °F	65	65	NA	25
Color	Brown-Black		Green-black	
Viscosity @ 100°F, cSt	2.27	9.56	---	10 max
Viscosity @ 100°F, SUS	34	57	NA	
Sulfur, wt%	0.20	0.30	0.44	1.30
Carbon Residue, wt%	1.29	2.33	7.42	0.4
Nitrogen, wt%	0.423	0.724	1.187	

Reference: (16)

TABLE 35. SYNTHOIL PROCESS PRODUCT SLATE

Product	Feed Coal, tons/day (moisture and ash free)			
	10,000	20,000	30,000	40,000
Naphtha, barrels/day	559.7	1,119.4	1,679	2,239
Middle distillate, barrels/day	176.6	353	530	706
Heavy fuel oil, barrels/day	25,060	50,119	75,178	100,238
Sulfur, tons/day	567	1,131	1,096	2,261
Ammonia, tons/day	47.8	45.5	143.3	191.1

Reference: (4)

TABLE 36. SYNTHOIL PRODUCT DISTILLATION DISTRIBUTION

<u>Component Fraction(%)</u>	<u>Crude Naphtha</u>	<u>Middle Distillate</u>	<u>Heavy Fuel Oil</u>
n-Butane	5.1	0	0.1
Pentane-400°F	88.4	2.6	0.1
400°-650°F	6.5	72.5	10.5
650°-850°F	0	19.9	17.1
850°-1050°F	0	5.0	25.0
1050°F	0	0	46.7
Residue	0	0	0.4
Ash	0	0	0.3

Reference: (4)

TABLE 37. HYDROCARBON COMPOSITION OF MIDDLE AND HEAVY DISTILLATES OF WEST VIRGINIA COAL FROM SYNTHOIL PROCESS

<u>Compound Types</u>	<u>Middle Distillate 207°-363°C</u>	<u>Heavy Distillate 373°-531°C</u>
Saturates	16.0	9.7
Mono-aromatics	27.3	4.7
Di-aromatics	21.6	22.6
Polyaromatics	7.9	41.1
Hetero-aromatics	22.2	15.6
Distillate Weight, % of syncrude	42.6	27.3

Reference: (16)

TABLE 38. SYNTHOIL MATERIAL BALANCE

<u>Input</u>		<u>Output</u>	
<u>Material</u>	<u>Weight Units</u>	<u>Material</u>	<u>Weight Units</u>
Coal	100.00	Exhaust gases	61
Water	422.10	Naphtha	0.59
		Middle distillate	0.21
		Fuel oil	34.84
		Sulfur	4.27
		Ammonia	0.04
		Slag slurry	55
		Cooling tower losses	366.15
TOTAL	522.10	TOTAL	522.10
Reference: (4)			

TABLE 39. THERMAL EFFICIENCY ANALYSIS FOR SYNTHOIL PROCESS

<u>Input Energy</u>	<u>Btu/ton Coal (as received)</u>
Coal input	22,400,000
Makeup hydrogen	8,200,000
Net process	600,000
Fuel fired	400,000
Coal pulverization	300,000
TOTAL	31,900,000
<u>Output Energy</u>	
Heavy oil product	19,700,000
Light oil product	1,200,000
Hydrocarbon gases	3,000,000
TOTAL	23,900,000
Thermal Efficiency = 74.9%	
Reference: (10)	

TABLE 40. SYNTHOIL ENERGY BALANCE

Input		Output	
<u>Material</u>	<u>Energy Units</u>	<u>Material</u>	<u>Energy Units</u>
Coal	100.00	Liquid product	59.6
		Sulfur, ammonia	1.7
		Wet cooling tower	27.0
		Air coolers	7.5
		Stack	<u>4.3</u>
		TOTAL	100.1

Thermal Efficiency = 59.6%

Reference: (3)

TABLE 41. SYNTHOIL CAPITAL REQUIREMENTS

	Feed Coal, tons/day (moisture and ash free)			
	1st Quarter 1978, 10 ⁶ dollars			
	<u>10,000</u>	<u>20,000</u>	<u>30,000</u>	<u>40,000</u>
Fixed capital investment	688.8	1158.5	1570.25	1948.4
Total depreciable plant	824	1386	1879	2331
Total capital required	899	1512	2049	2543

Reference: (4)

TABLE 42. SYNTHOIL ANNUAL OPERATING COST

	First Quarter 1978, 10 ⁶ Dollars			
	<u>10,000</u>	<u>20,000</u>	<u>30,000</u>	<u>40,000</u>
Materials and supplies	118.75	228.40	334.40	439.80
Labor	13.888	23.385	31.70	39.31
Property taxes and insurance	19.30	32.50	44.00	54.60
General and administrative overhead	<u>2.314</u>	<u>4.33</u>	<u>6.24</u>	<u>8.10</u>
Total	154.252	288.625	416.34	541.81

Reference: (3)

OTHER PROCESSES

Two additional liquefaction processes which have been developed, although less publicized than the foregoing processes, are TOSCOAL and CSF methods.

TOSCOAL

The Oil Shale Corporation (TOSCO), in addition to its oil shale technology efforts, piloted low-temperature carbonization of Wyoming sub-bituminous coals in its 23-ton/day pilot plant. The purpose was to upgrade the heating value in order to reduce transportation costs. The resulting char product represented 50 wt% of the starting coal, but retained 80 percent of its heating value. In addition, a low-sulfur oil could be recovered.

In this process (Figure 8), coal is fed from hoppers to a fluid-bed pre-heater, then to a pyrolysis drum where it comes in contact with ceramic

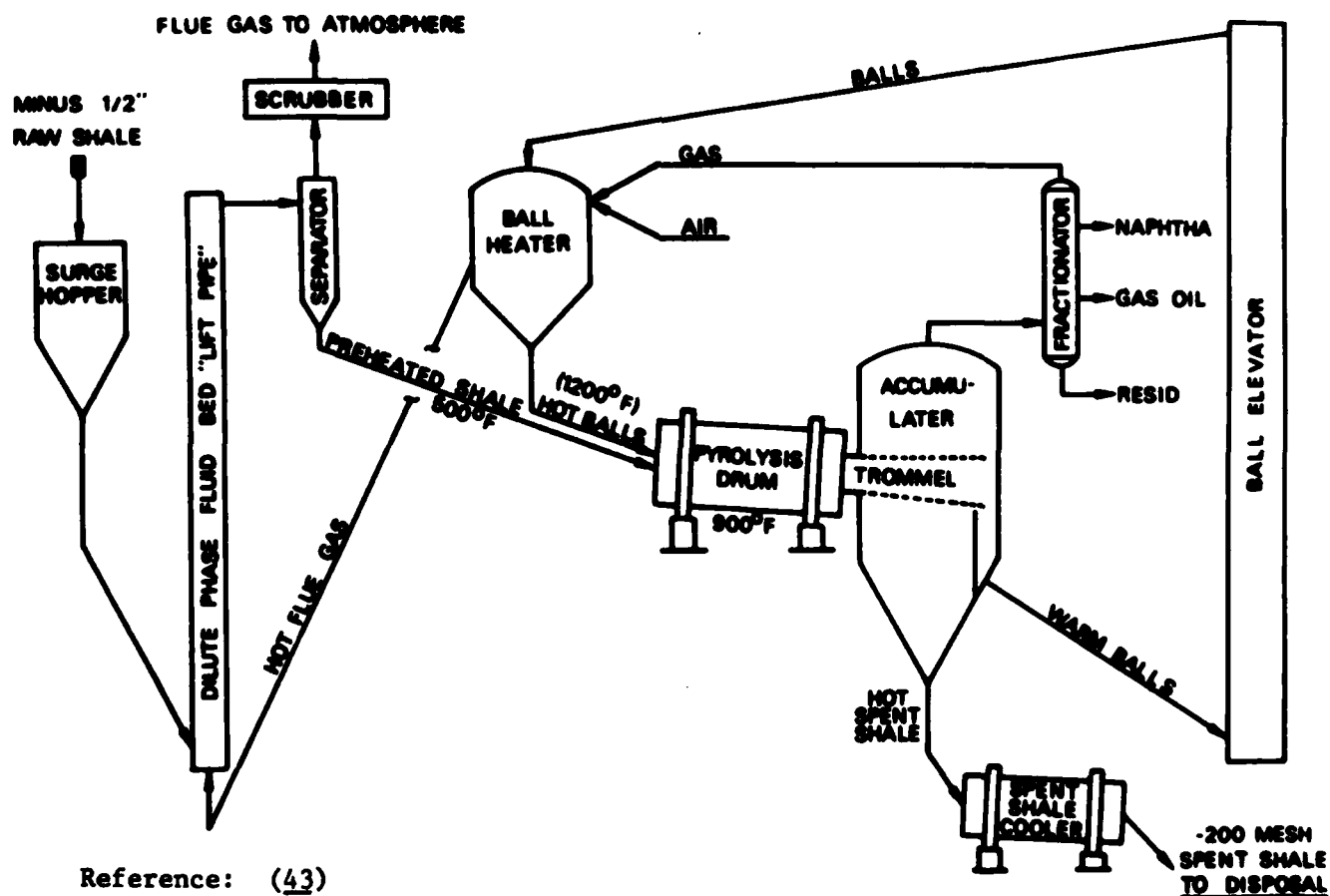


FIGURE 8. TOSCOAL PROCESS

balls heated by separated process gas. Drum effluent passes over a trommel screen to permit char product to fall through. Pyrolysis vapors are condensed and fractionated to produce oils. Light gases are also produced.

Properties of oils change only slightly with retorting temperature. They have high viscosity and cannot be transported through conventional pipelines. Heating values of product gases are comparable to those of natural gas, providing that C_4 compounds are included.

CSF

The Consol Synthetic Fuel Process (CSF) was developed by the Consolidation Coal Company, which initiated studies of liquefaction methods in the early 1950s. In the 1960s, molten zinc chloride was investigated as a catalyst, but regeneration and metallurgical problems were never resolved. In 1967, a 900 kg/hr pilot plant was constructed, but mechanical problems were encountered as well as declining catalyst activity and inability to demonstrate extract hydrogenation. Hydrocarbon Research managed to successfully treat the granulated extract, nonetheless.

The CSF process resembles the SRC process except that hydrogen is not introduced directly to the extraction reactor. Instead, it is used to rehydrogenate the extraction donor solvent after the extract is removed.

In the CSF process, coal is pulverized, slurried with a solvent, heated, and pressurized in a coil-furnace preheater. The finished product is then transferred to a staged extractor. The fundamental reaction is the depolymerization and solution of most of the coal in an aromatic solvent as a result of hydrogen transfer.

Since Consol's pilot facility never successfully developed, only estimates of product yield could be made. The liquid product would essentially be about 70 percent of the coal feed but with additional hydrogen to equal coal feed weight. Processing by Hydrocarbon Research converted about 70 percent of the extract actually obtained from the pilot plant to a distillate boiling at 200°-440°C. (2)

FOREIGN ACTIVITY

Four foreign nations which are highly dependent on foreign oil utilize or are interested in developing coal conversion technology. Their low-cost coal resources and large research and development budgets can permit them to accomplish their goals. These nations are South Africa, Japan, Germany, and Australia.

South Africa

SASOL ONE operated inefficiently during the 1950s, basically because a small pilot plant was scaled-up to a commercial-size venture without an intermediate stage to prove the viability of the process on a larger scale. The result was that design of process components which allowed the operation to work on a small scale presented problems when applied to a large, commercial-size operation.

The lessons learned from SASOL ONE were heeded for the planning of SASOL TWO and THREE. The resultant higher efficiency and lower cost of operation will result in a combined output of 100,000 barrels of motor fuel per day. One unit of great importance in the process is that SASOL TWO and THREE will use Synthol fluid bed reactors which have larger capacity than the Arge fixed-bed type. SASOL TWO and THREE will use multiple Synthol reactors only instead of a combination of both types used in SASOL ONE, permitting considerable savings in capital costs.

Japan

Japan depends on imported oil for 75 percent of its energy requirements. Rate of importation is increasing, but R&D efforts into alternative energy development give hope that this rate will stop by 1985.

There are no detailed indications or plans of Japan's coal technology development other than basic research into alternative energy sources. Of the budget allocated for this research, 70 percent will be for coal conversion

processes. It is projected that coal will supply most of Japan's energy needs by 1995.

Germany

The development of coal industry in Germany for the future looks grim for two reasons: the lack of domestic coal which would mean that the development of coal liquefaction would depend largely on imported coal, and the high cost of German hard coal which costs about twice as much as imported coal. Thus for some time to come, Germany will almost always find it more economical to import oil. Nonetheless, research and development is being done through government support, particularly by joint venture of Ruhrkohle and Veba Oil utilizing the new IG (Farben) process.

By the late 1980s, coal conversion is expected to be somewhat more competitive with foreign oil so the motive for development is both import substitution and export of technology and expertise to pay for imports.

A joint venture has been undertaken with the Australian government to investigate the potential of converting brown coal (lignite) to liquids from which diesel fuel might be made. Most of the private funding is from Australia, and government funding is from Germany. The purpose is to examine thermal efficiency and economic parameters to find the proper combination of input materials to convert coal to motor fuels.

Australia

Australia is currently about 65 percent self-sufficient for its oil needs. It is expected to remain between 40 and 80 percent between now and the year 2000, but will depend on the discovery of new oil. Australia is also a net exporter of coal, and its coal industry is among the world's most efficient and thus has been able to compete in the European market.

Since coal deposits are in the Australian interior, considerable investment would be required for the nation to be a major producer and exporter. The

expansion of exportation could be hindered by the problem of agreement between exporters on investment financing methods. The government prefers to seek funding through foreign investment.

In addition to ventures with Germany, the Nippon Brown Coal Liquefaction Co. Ltd. project has been organized by Australia and Japan. It was originally proposed for solvent-refined coal, but has been redirected to production of liquid fuels. The project is managed entirely by Japanese companies.

Nippon project research into SRC solid production has been completed, but work is needed on the upgrading of unreacted coal to produce liquid. A feasibility study has led to construction plans for a pilot plant followed by a demonstration plant, and eventually, a commercial plant, providing satisfactory results are obtained. Australia has agreed to contribute land, coal, and some labor to the pilot phase. Pending decisions by the Australian government, production and commercial operation will be shared between Japan and Australia.(5, 17)

PROBLEMS WITH COAL LIQUID TECHNOLOGY

Conversion Processes

COED

The char product from the COED process contains too much sulfur to be advantageous to consumer power companies, although the product oil and gas will probably be marketable.(1)

EDS

The pressure drop across the lead solvent hydrogenation reactor can be minimized by care in controlling the recycle solvent endpoint, but the coal conversion remains lower than expected.(18) There are, however, several possible ways of increasing liquid yield; recycling of coal liquefaction bottoms to increase their residence time in the reactor, increasing the solvent/coal ratio by 50 percent to provide increased donor hydrogen availability, and increasing overall thermal efficiency of an energy-balanced plant.(10)

Exxon has identified several problems for raw distillate EDS fuel utilization:

1. EDS fuel oil is heavier than water, unlike petroleum. This impacts fuel storage and handling systems.
2. NO_x emissions are higher than those from petroleum fuels, which is consistent with the higher fuel-bound nitrogen content.
3. EDS fuel oil is incompatible with petroleum fuels due to presence of gas oil and coker liquids. Separate storage and handling features would be necessary for a plant using both EDS and petroleum. Additional upgrading will be necessary.(19)

H-Coal

When producing heavy oil, separation of heavy liquids from solids (such as ash) by distillation presents difficulty. A de-ashing unit may be acceptable.(18)

SRC-I

The main process problem is separation of solids, particularly ash, from the slurry.(5)

SRC-II

A considerable design problem exists in balancing temperature, residence time, and coal concentration during the slurry mixing operation. When recycle slurry is mixed with fresh pulverized coal, high temperature is required for heat economy; however, the viscosity of the coal/slurry mixture increases with temperature and coal concentration. This is due to the formation of a gel which is difficult to mix and pump.(13) It also presents an additional problem in transit from the slurry preheater in that maximum practical tube diameter is limited by heat transfer, and thus requires multiple tube passes.(10)

Naphtha and fuel oil are not directly interchangeable with their petroleum counterparts due to differences in chemical composition. Consumer acceptance will require different methods of use. SRC-II boiler fuel handling properties equal those of petroleum but common elastomers except Viton, Nylon 616, and Teflon are degraded after prolonged contact with the coal oil. Both boiler and turbine fuel oils have a higher nitrogen content than petroleum fuels and yield higher NO_x emissions. Economic removal of nitrogen must be explored, and alternate methods of reducing NO_x are needed. Also, turbine oil has shown erosion of turbine blades.(13)

The aromatics content of SRC-II middle distillate is high, rendering it unsatisfactory for diesel engine performance and for use as home heating oil. Severe hydrotreating is needed.

The cost of hydrogenation for SRC-II is 35.7 percent, compared to 19 percent for SRC-I. The cost of distillate upgrading would run from \$8/bbl for gasoline and heating oil to \$13/bbl for mostly gasoline. Also, a substantial investment would be required to manufacture ethylene from ethane and propane.(5)

Hydrocarbonization

Handling of coking coals has proven to be a serious obstacle in process development and was one of the principal contributing factors to the failure of the Coalcon project. Since that time, several successful approaches have been demonstrated. Among these approaches are reactor configurations and chemical pretreatment of coal.

Utilization

The hydrocarbonization low-sulfur product char can be used as a boiler fuel or for the production of metallurgical coke. High-sulfur char can be used in a fluidized-bed combustor or as a boiler fuel in a conventional furnace with flue gas desulfurization; however, its use as a boiler fuel is not economical at present. Research continues on optimization of the hydrocarbonization process for the elimination of char production.(10)

Basically, the problems with coal-derived diesel fuels are as follows:(19)

Problem

Poor combustion, smoking
Excessive cylinder wear
Injector nozzle plugging
Injector pump sticking
Excessive engine deposits

Probable Cause

Low cetane number
High sulfur levels
High boiling fraction impurities
High viscosity
High sulfur/heteroatom content

Other anticipated problems are the short time available for combustion, ignition delay, and hot corrosion.(19)

Coal-based distillates have a high level of aromatics when compared to petroleum distillates. This has an impact upon utilization as diesel or turbine fuels. They are inadequate as diesel fuels due to resultant high viscosity and low cetane number. They make inferior turbine fuels because of their high freeze point and low smoke point.(19) Generally, coal products from direct liquefaction contain more aromatics and high molecular weight aliphatics and thus have a higher octane number than fuels derived from indirect liquefaction which contain smaller alkanes and favorable cetane number. Performance test results of coal-derived diesel fuels on a CLR engine were compared with those of a petroleum-based diesel. It was found that fuel consumption varied from 96 to 105 percent of the base fuel, particulate emissions ranged from 1.6 to 3.1, hydrocarbon emissions 1.2 to 2.1, and nitrogen oxides emissions from 1.0 to 1.1 times that of the base fuel. After 115 hours of operation, low but measurable wear on cylinders and piston rings was noted. Lubricating oil showed increased iron content and carbon residue, indicating relatively low lubricity and thermal stability of coal-derived diesel fuel. Vehicle tests employing coal-derived diesel showed anticipated fuel economy and particulate emissions similar to that of the base diesel. However, unburned hydrocarbon emissions were lower and nitrogen oxides emissions higher for the coal fuel.(22)

It is technologically possible to make the utilization of coal-derived diesel fuels a reality, but the development of new engines will be required. One potentially feasible type is the direct-injected stratified charge (DISC) engine. Also, chemical treatments to stabilize coal-derived fuel and a method to control NO_x emissions (possibly by solvent extraction of nitrogen compounds) are desired.(20)

Properties critical to diesel fuel performance and exhaust emissions such as cetane number, viscosity, soot formation, heating value, and volatility are dependent upon fuel composition. An "ideal" diesel fuel has never existed, but it is known that engine performance and emissions improve with fuel's higher alkane and lower naphthenic and aromatic content. Petroleum diesel fuel has never been upgraded to approximate an "ideal" composition. This possibility offers a challenge to fuels engineering in synthetic fuels

technology as an area of research to optimize performance of coal-derived distillates. The proper composition would have to consist mainly of long straight-chain paraffins, little or no isoparaffins or naphthenes, and virtually no aromatics. Volatility may increase slightly, but cetane number, viscosity, heating value, exhaust emissions, and toxicity would improve.

It is difficult to produce specification aviation turbine fuels from catalytic hydroprocessing of coal liquid fractions for two reasons. First, coal liquid fractions are high in aromatics and low in paraffins; second, typical hydroprocessing catalysts convert aromatics to naphthenes rather than paraffins. Coal liquids have low paraffins at all hydroprocessing severities; thus, regardless of how the liquid is refined, it will always have higher density and lower flame radiation, freeze point, and weight heat of combustion than petroleum jet fuels.(23)

Storage Stability

While in storage, fuels tend to deteriorate over time. Products of deterioration are potentially harmful to engine operation in that they can cause fuel filter plugging, lower efficiency and fuel economy, and corrosion.(6)

Since coal liquids are candidate future fuels, their storage stability must be assessed and corrected in a manner similar to that for petroleum fuels. In one study, accelerated aging of coal liquid at elevated temperatures has shown that viscosity increases proportionately to temperature, time, and richness of oxygen atmosphere. With viscosity increase came increases in degradation product concentration. These products consisted of higher molecular weight components than were present in the unaged liquid. Infrared analysis of the aged samples showed the formation of new spectral bands indicative of C=O and C-O groups. Analysis of the filtered liquid portions of samples aged in an oxygen atmosphere showed an increase in weight percent saturates and a decrease in weight percent aromatics, acid, bases, and resins as well as a decrease in oxidation rate with time.

A striking feature was noted during the study. The coal-liquid samples under investigation had appreciable phenol concentrations, yet the liquids deteriorated in the presence of these assumed potential inhibitors. One possible explanation is that the naturally present phenols are of different structure than phenols in typical antioxidants. To check this possibility, the acid fractions of the samples were stripped of their -OH functionalities by trimethylsilyl derivatization. These fractions subsequently showed essentially no oxygen consumption, which may mean that the natural phenols, rather than acting as inhibitors, either act as the specific sites for the oxidation process or are involved in propagation steps of the gum-formation mechanism.(24)

Environmental, Health and Safety

Environmental

The use of materials such as coal which are burned to produce heat can have potentially adverse effects on all sectors of the environment. Upon burning, gaseous emissions occur which contain sulfur and nitrogen oxide, CO, CO₂, and various organic compounds which could have toxic effects on animals and plants. Particulate matter in the stack emissions, as well as the unburned residue, both of which are higher in metal concentration than the original coal, must be disposed of. Finally, liquid effluents, primarily water, must also be contained. This includes not only process water, but rain water which will leach the ash dump facility. Similar problems are met in mine areas and alternate fuel processing facilities. In addition, mining areas are noted for removal of vegetation, destruction of soil structure, loss of wildlife habitats, changes in hydrology, and severe erosion and sedimentation.(25)

Health and Safety

Health hazards can range from chronic illness to premature or immediate death. Genetic defects may also arise, thus affecting future generations.

It is considered that stack emissions are probably the greatest source of toxic materials for human exposure. National Air Quality Standards were implemented to control levels of SO_x , NO_2 , particulate matter, CO, photochemical oxidants (e.g., O_3) and nonmethane hydrocarbons in the atmosphere. Other materials may, however, be of greater health danger than those listed. The importance of emissions are shown in studies of acute exposures in certain areas which have caused excess mortality.

Coal mining itself can lead to bronchitis, coal workers' pneumoconiosis, and other diseases through inhalation of the dust created. The U. S. Coal Mine and Safety Acts of 1969 and 1977 limited dust levels in mines with the intent of reducing pulmonary diseases in miners. An idea of the extent of the disease is reflected in payments of benefits to miners from 1970 and 1973 which equalled 1.7 billion dollars.(25)

METHODS OF ANALYSIS AND CHARACTERIZATION

Elemental Analysis

The determination of oxygen content of coal liquids by neutron activation analysis eliminates the error of determination by difference. The method measures oxygen with assistance of a calibration curve preconstructed from analysis of standard oxygenates.(26) Carbon, hydrogen, sulfur, and nitrogen can be determined using standard methods providing the physical character of the sample does not prohibit their use.(26)

In a study of coal liquid methods of analysis, sulfur was determined by oxygen bomb combustion, silicon by gravimetric digestion procedure, and aluminum, iron, magnesium, calcium, and potassium by digestion/atomic absorption spectroscopy. Trace elements of copper, chromium, nickel, manganese and zinc were determined by silica-free ash fusion and atomic absorption. Lead and cadmium were complexed with iodide, then extracted with methyl-isobutylketone prior to atomic absorption analysis.(27)

Most of these elements can be determined in coal or coal liquids by combustion of the sample in an oxygen bomb containing water. The elements will dissolve in the water and can be measured using conventional techniques.(28)

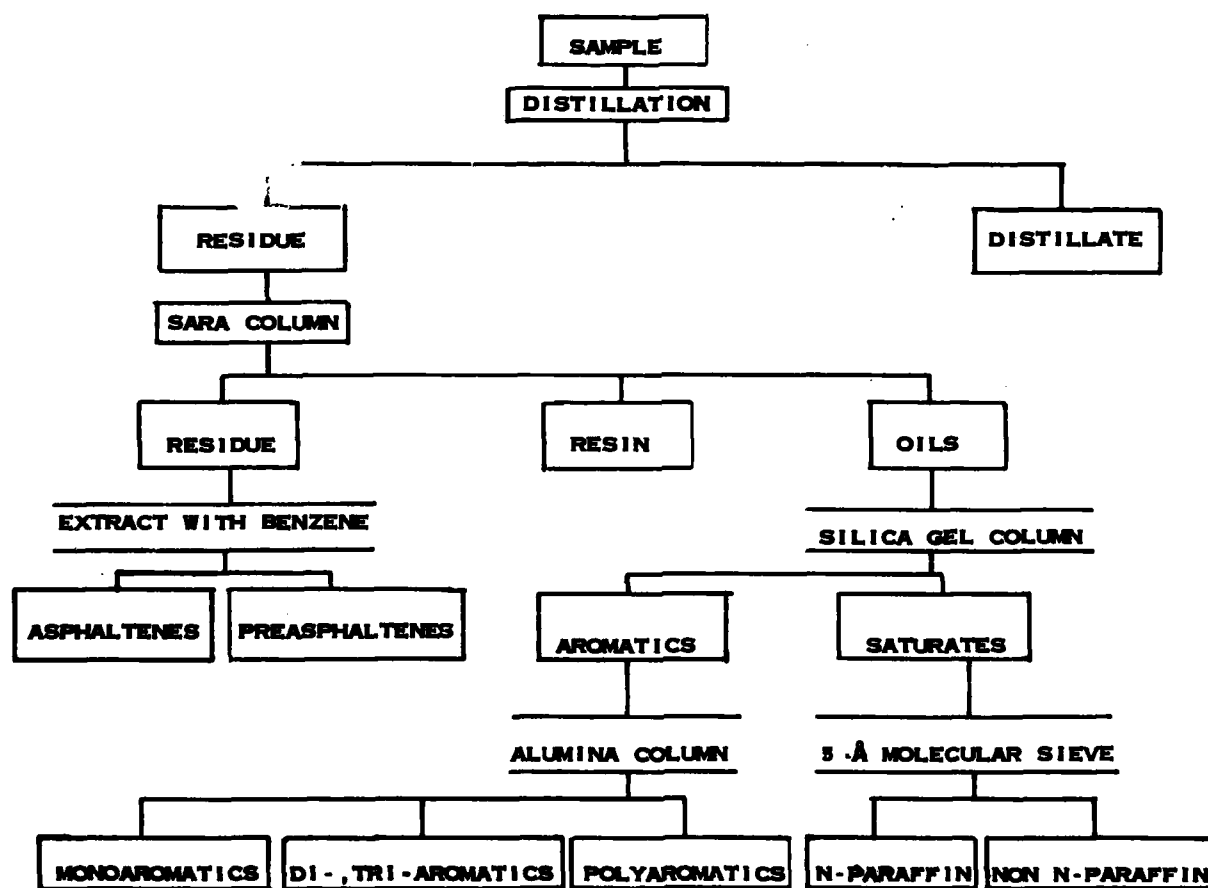
Compositional Analysis

Gas chromatography/mass spectrometry is useful for the determination of some nitrogen compounds. The gas chromatograph can separate compounds with similar molecular weights within the same class virtually indistinguishable on a mass spectrometer (e.g., 6-methylquinoline and 7-methylquinoline). The mass spectrometer can be valuable in establishing measurements in the presence of impurities with identical or nearly identical retention times as the subject compounds.(29)

Nitrogen bases (pyridines, anilines, and quinolines) in light coal oil can be precipitated by hydrogen chloride gas, addition of sodium hydroxide and

ether, washed with benzene and analyzed by GC/MS. The method has the advantage of no side reactions with hydrogen chloride.

For general characterization of a coal liquid sample, the sample can be separated in accordance with Figure 9. Carbon and hydrogen can be determined by microcombustion and other elements determined by conventional means for the separated distillate, resins, asphaltenes, preasphaltenes, aromatics, n-paraffins, and non n-paraffins. Conventional methods of analysis include molecular weight by osmometry or ^1H NMR, nitrogen by Kjeldahl method, oxygen by neutron activation, sulfur by combustion method similar to ASTM D 1552, distillation by ASTM D 2887, and analysis of saturate fractions by MS.(30)



Reference: (31)

FIGURE 9. SCHEME FOR SEPARATION OF COAL LIQUID FRACTION

Alkylphenols can be determined in coal-derived solvents by silica gel column separation followed by HPLC separation. The resulting fractions are characterized by base extraction, chemical spot tests, ultraviolet absorption, and fluorescence spectrometry. The phenols are identified by final HPLC.(31)

Grade 923 silica gel and grade 62 silica gel impregnated with trinitrophenol can aid in separation of some heavy oil coal liquid components according to their number of condensed rings. The greater the number of condensed rings in a compound, the less strongly it is adsorbed by the packing. This is good for separation of PAHs. Phenanthrene, pyrene, coronene, and chrysene can easily separate alone and in solution with nonaromatic components.

A method of separating components of solvent-refined coal is by dissolution with tetrahydrofuran with subsequent molecular separation on Bio-Bead columns. (Bio-Rad Labs, Rockville Center, NY) The bead mesh facilitates control of molecular size and can help to prepare as many fractions as desired.(32)

Elution behavior of components of coal-derived liquids can be demonstrated on silica gel. Order of elution is saturate, hydroaromatic, polynuclear aromatic, phenols, nitrogen bases. Thin-layer chromatography and reactions involved can give a semi-quantitative measure of the degree of hydrogenation of coal-derived liquids.(32)

Benzene solubles (asphaltenes) and pyridine solubles (preasphaltenes) are quantitatively determined gravimetrically by evaporation and precipitation or by filtration by Buchner funnel. Solid/liquid separation is effected by centrifugation and filtration.(33).

Separation of coal liquid oils, asphaltenes, and residue can be accomplished with ion-exchange and silica gel chromatography. Ion exchange chromatography fractionates oils and asphaltenes into acids, base, and neutrals while silica gel chromatography separates neutral fractions into saturate and aromatic concentrates. Functional groups are then identified using infrared spectrophotometry and carbon-number distributions by mass spectrometry.(34)

Procedure descriptions exist for methods of determining oils, asphaltenes, and insolubles. Methods also exist for evaluation of production streams from liquefaction development process units, a rapid means of analyzing a small number of samples, and as a means of studying conditions for the precipitation of asphaltenes and preparation of fractions for further study.(36)

A detailed method for the determination of asphaltenes describing apparatus reagents and procedure is given by Schultz.(37)

Physical Properties

Enthalpy of coal liquids can be determined by bomb calorimeter. A study of various enthalpy equations applied to coal and petroleum for comparative purposes yielded various values for each. The study also included elemental and hydrocarbon characterization of coal and petroleum liquids.(38)

The heating value of a coal liquid can also be calculated from its percent of C, H, N, O, and S. Differences in calculated and measured values vary from 0.16 percent to 2.86 percent.(39)

The physical properties of any liquid fuel are to some extent dependent upon the fuel's composition. Freezing point is related to average aromatic diameter, heat of combustion to hydrogen content, smoke point to aromatic concentration, and viscosity also to average aromatic diameter.(40)

There is also evidence of viscosity related to asphaltene content in coal-derived liquid. Asphaltenes in coal liquid are aromatic compounds. Since the logarithmic viscosity number is related to the degree of intermolecular interaction between the polar functional groups of the asphaltene, it may be that the degree of hydroprocessing in coal-to-liquid conversion can have a bearing on the viscosity of coal liquids.(41)

An investigation of the effect of toluene insolubles and asphaltene acid, basic and neutral fractions of coal liquid on viscosity was undertaken. The conclusions were:(42)

1. The natural logarithm of the viscosity ratio (η / η_0) shows linear dependence on the concentrations of these solutes.
2. The toluene insolubles, on a weight basis, produce a viscosity twice that produced by the whole asphaltene fraction.
3. The basic asphaltene subfraction has a greater effect on viscosity than does either the acid or neutral subfraction of the whole asphaltene fraction.
4. Combinations of various weight fractions of basic, acid, and neutral asphaltene subfractions have a synergistic effect on viscosity.
5. The aggregation of asphaltenes and toluene insolubles plays a significant role in the viscosity of coal-derived liquids.

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COAL LIQUIDS: MANUFACTURE AND PROPERTIES A REVIEW(U)
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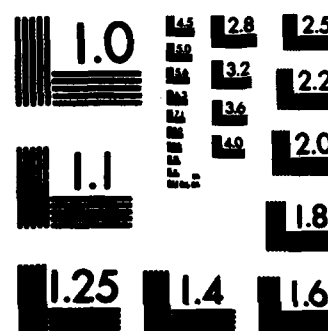
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